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GULF ENVIRONMENTAL SYSTEMS

GA-9860 (Rev.)
Addendum B

FINAL REPORT OF REVERSE OSMOSIS APPRAISAL PROGRAM AT TOM'S COVE, ASSATEAGUE ISLAND NATIONAL SEASHORE, VIRGINIA

by

R. L. Truby

Prepared under
Contract No. 14-10-7:991-039
for the
National Park Service
U.S. Department of the Interior

January 15, 1971

GULF ENVIRONMENTAL SYSTEMS COMPANY
P.O. BOX 608, SAN DIEGO, CALIFORNIA 92112

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ABSTRACT

A reverse osmosis field demonstration has been conducted in the Tom's Cove area, Assateague Island National Seashore, Virginia. The purpose of this demonstration was to illustrate the ability of reverse osmosis to produce potable water from the local brackish well water. A trailer-mounted reverse osmosis unit capable of producing 4000 gpd of product water was utilized for this demonstration.

After preliminary delays, a continuous 11-day demonstration run was carried out. During this period the reverse osmosis membrane modules consistently rejected 92.9% of the salts in the well water. Owing to the high dissolved solids content of this water (approximately 19,000 mg/l), 92.9% rejection was not sufficient to produce potable water. Consequently, 100 gal of product water was collected and passed through the unit a second time, simulating a two-stage system. The product water from the second pass was of excellent quality, and with slight pH adjustment would have been potable.

In this report, the results of the field demonstration are presented and the economics of a series of reverse osmosis units are discussed. At Tom's Cove a two-stage reverse osmosis unit would be required to produce potable water from the local well water. The cost of potable water produced by a two-stage 100,000-gpd reverse osmosis unit using standard Model 4000 modules would be $2.40/1000 gal. The cost of potable water produced by a two-stage 100,000-gpd reverse osmosis unit using high-flux modules would be $1.84/1000 gal.

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1. INTRODUCTION

A series of reverse osmosis[*] field demonstrations has been carried out by Gulf Environmental Systems (GES) for the National Park Service, U.S. Department of the Interior. The results of two previous field demonstrations have already been reported.[**] This report describes the results of a field demonstration conducted at Tom's Cove, Assateague Island National Seashore, Virginia, under Contract No. 14-10-7:991-039.

The main purpose of this field demonstration was to illustrate the ability of reverse osmosis to convert a brackish well water supply into a potable water source. Utilization of this process offers several advantages over other potential methods of acquiring potable water, i.e., importation by truck or pipeline. Some of the more important advantages are:

1. Maximum Utilization and Conservation of Available Water
   Importation of potable water by truck or pipeline offers an immediate solution to the water shortage problem in any area. However, as the demand increases, the supply of untreated potable water can be expected to rapidly disappear. Reverse osmosis offers the capability of converting substandard brackish water supplies, and water supplies declared unsafe by the U.S. Public Health Service,[***] into immediately available potable sources. In addition, reverse

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[*] See Appendix A for a description of reverse osmosis.

[**] Poole, G. F., and R. G. Sudak, "Final Report of Reverse Osmosis Appraisal Programs at Chaco Canyon, New Mexico, and Petrified Forest, Arizona," National Park Service, U.S. Department of the Interior, Report GA-9860 (Rev.), Gulf General Atomic, December 1969; and Truby, R. L., "Final Report of Reverse Osmosis Appraisal Programs at Furnace Creek and Stove Pipe Wells in Death Valley National Monument," National Park Service, U.S. Department of the Interior, Report GA-9860 (Rev.), Addenum, Gulf General Atomic, June 11, 1970.

[***] See Appendix B for U.S. Public Health Service Standards.

osmosis can be applied to already potable water sources in such a way as to reduce problems such as excessive concentrations of sodium chloride, calcium carbonate, iron, and other solutes. Thus, the net effect of using reverse osmosis would be maximum utilization of the total water supply in an area.

2. Flexibility for Expansion

   The use of reverse osmosis offers flexibility in that the amount of potable water produced by a unit can easily be expanded in response to the demand for it. This is, of course, subject to the limitation of the raw water supply. The increase in the production of a unit is accomplished merely by adding the necessary number of ROGA® modules (each capable of producing 300 gpd of product water*) and increasing the pump capacity. This flexibility for expansion has the obvious advantage of allowing areas of accelerated growth to respond rapidly to the public demand.

The above considerations are in accord with the National Park Service in its efforts to serve the public and protect the environment. The field demonstration described in this report supplies actual field data illustrating the ability of reverse osmosis to convert a brackish water supply into a potable water source.

The field demonstration conducted at Tom's Cove was carried out from June 14 until July 5, 1970. During this period data from more than 11 days of continuous operation were gathered. The results of these tests and the economics of a range of reverse osmosis units capable of treating the brackish water at Tom's Cove are the major concern of this report.

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*The rated output of each module has been adjusted to allow for the high-concentration well water found at Tom's Cove.

2. DESCRIPTION OF DEMONSTRATION UNIT

The trailer-mounted 4000-gpd reverse osmosis unit utilized for the demonstration at Tom's Cove is shown in Fig. 1. Its components and size rating follow the same standards and guidelines as all commercial and Government reverse osmosis units sold by GES. This commonality serves the dual purpose of providing interested parties with a first-hand look at reverse osmosis and at the same time permitting GES to realistically evaluate the performance of a unit operating on the local water.

The demonstration unit consisted of pretreatment equipment, high-pressure pumps, and nine spiral-wound modules contained in three pressure vessels. These components, which are described below, were mounted to the floor of a flatbed trailer for easy transport and display.

PRETREATMENT ASSEMBLY

The pretreatment assembly consisted of two chemical addition pumps, two full-capacity cartridge filter housings and filters capable of removing particulate matter larger than 5 μ from the feedwater, a pH controller, and a 50-gal tank, complete with mixer, for addition of extra chemicals. One of the chemical addition pumps was automatically operated by the pH controller and added the correct amount of concentrated sulphuric acid (66° Baumé) to maintain the feedwater pH between 5.0 and 6.0. The second chemical addition pump was used for injection of chlorine and/or sodium hexametaphosphate into the feedwater. When one or both of these chemicals were added, they were mixed and stored in the 50-gal tank.

HIGH-PRESSURE PUMPS

After pretreatment the feedwater was pressurized by two in-series, REDA, submersible centrifugal pumps. These pumps required 230-V, single-

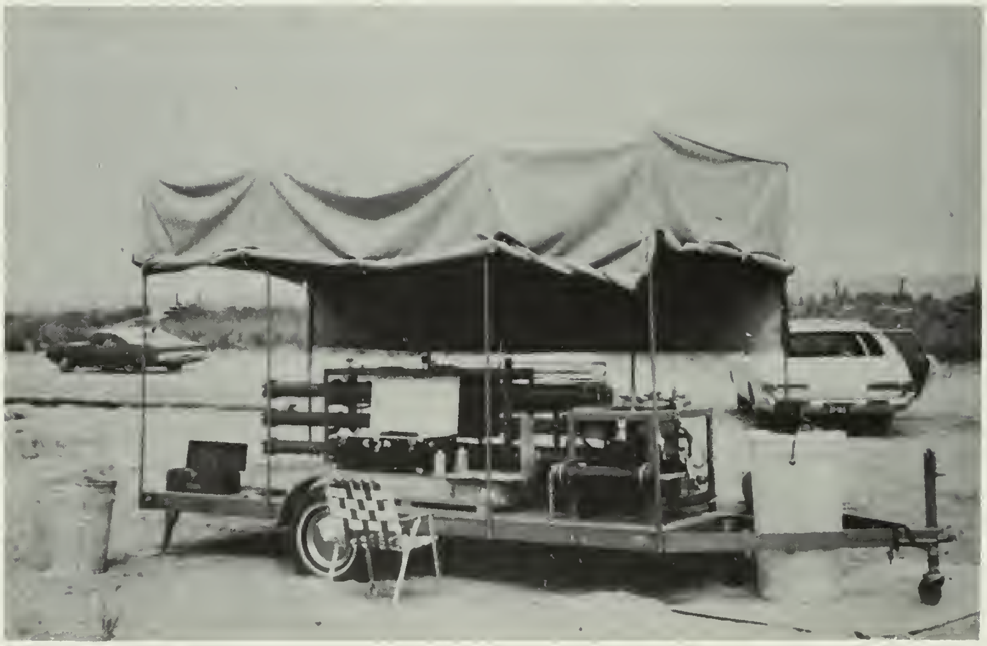

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Fig. 1. 4000-gpd reverse osmosis demonstration unit phase power and were capable of delivering 6 gpm at a pressure of 500 psi. This type of pump characteristically operates at around 45% efficiency, and consequently draws excess power. The pumps are used with the trailer unit because they are small and can quickly be replaced in case of failure. Positive displacement pumps that operate at around 85% efficiency are usually supplied as part of a permanent installation.

MEMBRANE-MODULE PRESSURE VESSEL ASSEMBLY

After pressurization the feedwater was manifolded into the first two pressure vessels, which were in parallel. The brine from these two vessels was collected and used as the feedwater for the third vessel. The product water from all three vessels was combined and discharged through one hose, while the brine was discharged through another. A schematic diagram of the entire system is shown in Fig. 2.

The modules contained in the three pressure vessels were standard Model 4000 ROGA spiral-wound modules, each having a nominal membrane area of 50 ft^2. The spiral-wound module is shown diagrammatically in Figs. 3 and 4, while Fig. 5 shows the modules installed in a pressure vessel.

The net driving pressure at which a unit is operated, the membrane area, the feedwater temperature, and the flow scheme combine to determine the quantity of product water produced. The field demonstration unit was designed to produce 4000 gpd of product water at 77°F, a net driving pressure of approximately 500 psi, and 50% recovery (one out of 2 gal of feedwater recovered as product water).

The recovery of this unit was limited by the exit brine flow, which should be in excess of 2.5 gpm to lessen the effects of concentration polarization. Concentration polarization occurs when a layer of concentrated dissolved and suspended materials builds up at the membrane surface as the product water is removed. The brine flow through 50-ft^2 modules must be kept in excess of 2.5 gpm to flush this foreign material from the unit.

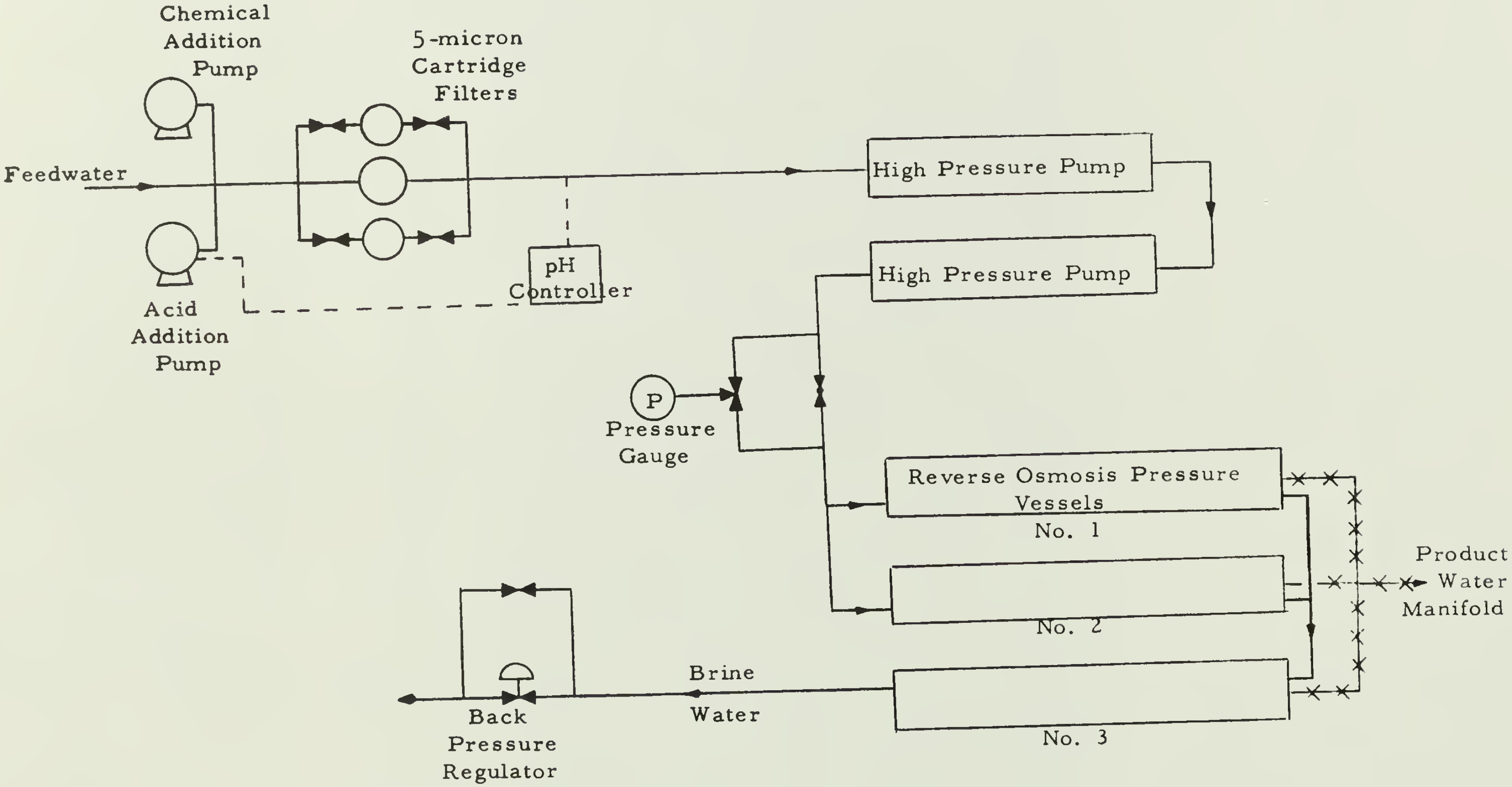
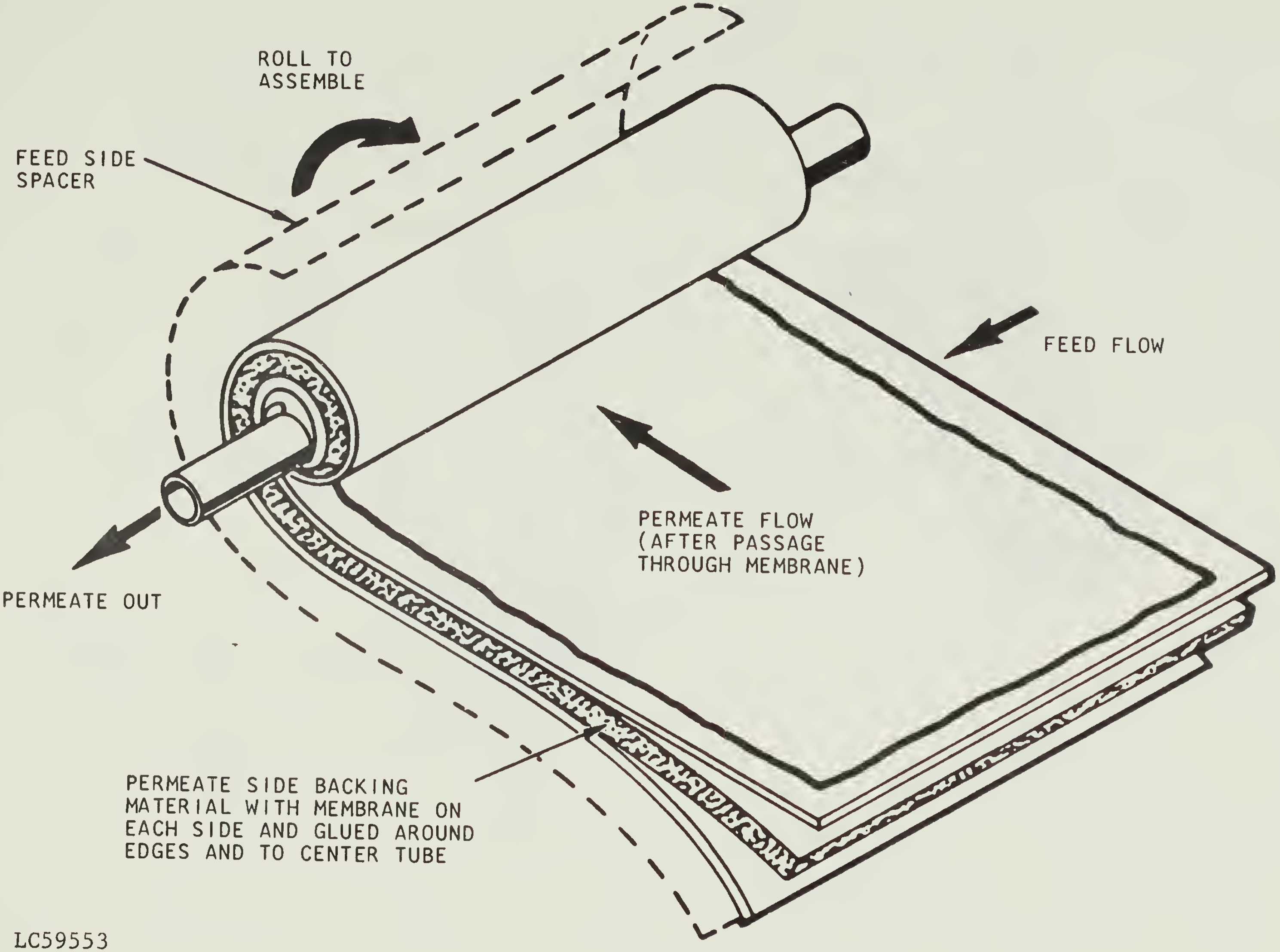


Fig. 2. Schematic of reverse osmosis demonstration unit


ROLL TO ASSEMBLE
FEED SIDE SPACER
FEED FLOW
PERMEATE FLOW (AFTER PASSAGE THROUGH MEMBRANE)
PERMEATE OUT
PERMEATE SIDE BACKING MATERIAL WITH MEMBRANE ON EACH SIDE AND GLUED AROUND EDGES AND TO CENTER TUBE

LC59553

Fig. 3. Spiral-wound module concept

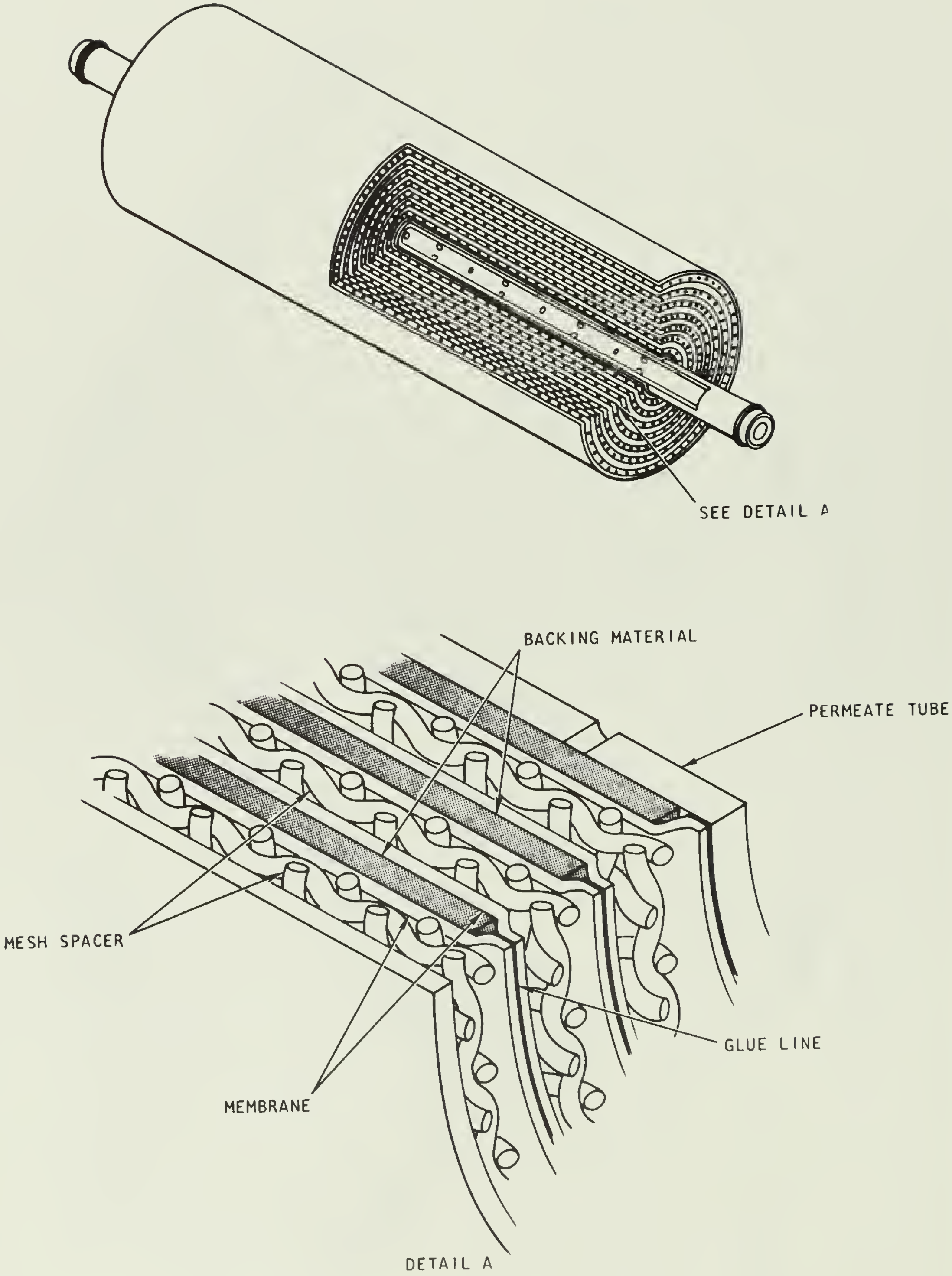


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Fig. 4. Spiral-wound module cutaway

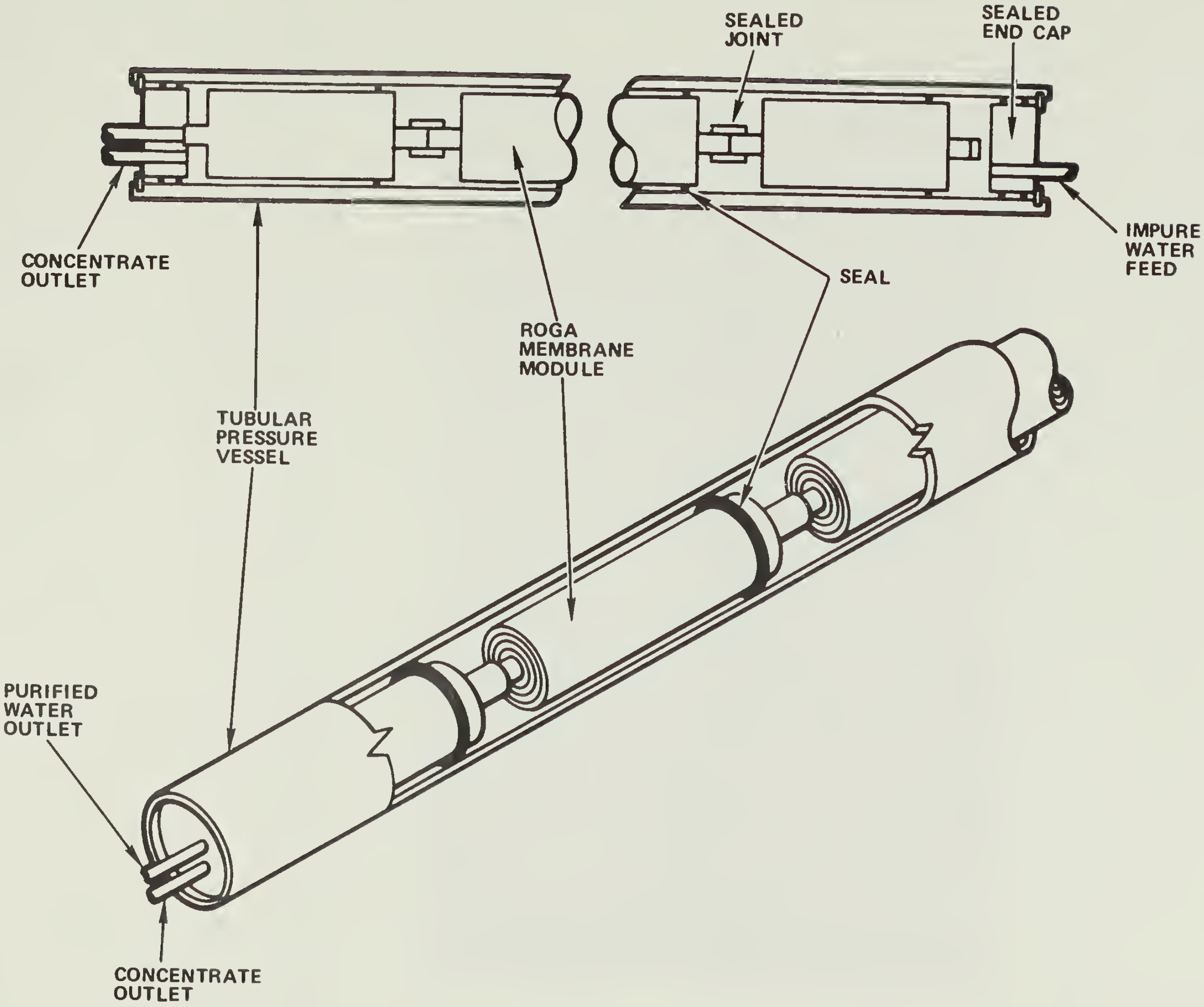

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Fig. 5. Module orientation in pressure vessel

3. PERFORMANCE OF UNIT AT TOM'S COVE

The demonstration unit, accompanied by a GES representative, arrived at Tom's Cove on June 14, 1970. It was installed at a site in the beach area near a brackish well (see Fig. 6).

From June 15 until June 24 unit operation could be conducted only on an irregular basis due to various mechanical and electrical problems. During transcontinental transport the unit had been damaged, and minor repairs were required before testing could begin. In addition, the well proved to have insufficient capacity, and after it had successfully been enlarged, the well pump failed.

By June 25 the majority of the problems were solved. At this time 24 hr per day operation was initiated and maintained for 11 consecutive days. During this period a total of 68,280 gal of brackish well water was treated; 19,560 gal was recovered as product water, leaving 48,720 gal of concentrate. This is equivalent to an average recovery of 28.6% (approximately 3 out of every 10 gal recovered as product water) for the entire run.

From June 25 until June 30 the unit operated at 500-psi pump pressure. The recovery during this period was 24.5%. During July 1 through July 5 the unit pressure was increased to 600 psi, and the recovery was subsequently raised to 33.8%.

EFFECT OF OSMOTIC PRESSURE ON RECOVERY

The main factor affecting the recovery level of the unit during the demonstration run was the osmotic pressure of the raw well water. For sodium chloride solutions the osmotic pressure can be calculated as follows:

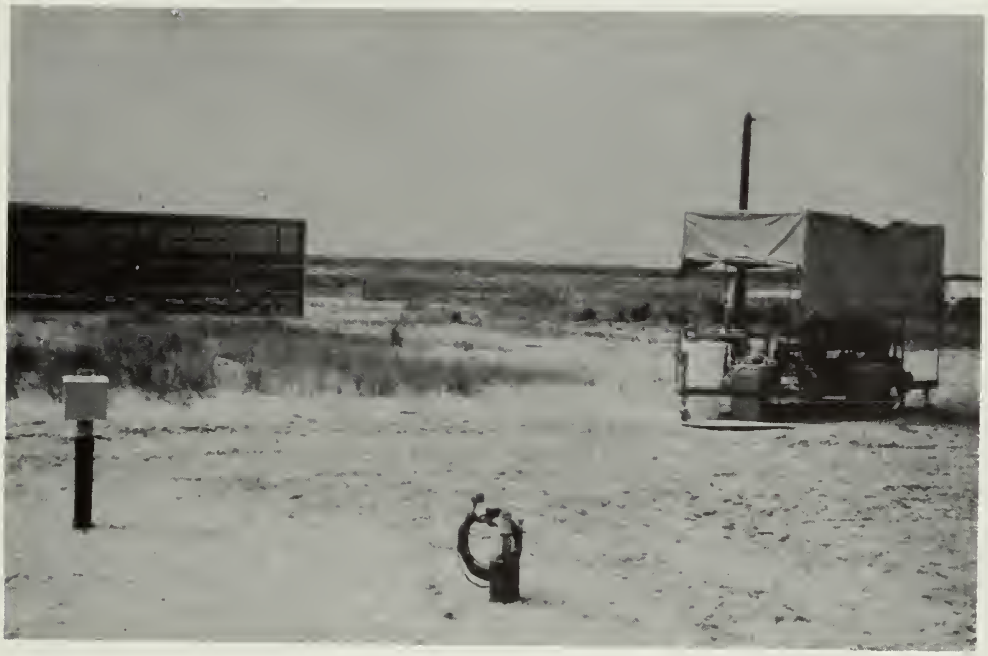

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Fig. 6. Reverse osmosis demonstration unit at Tom's Cove, Assateague Island National Seashore. Brackish-well head appears in foreground.

$$\text{Osmotic pressure } (\Delta\pi) = 0.0115 \times \text{mg/l sodium chloride},$$

where 0.0115 is derived from the Van't Hoff equation. Since the total dissolved solids (TDS) content of the raw well water is almost 75% sodium chloride, an approximation of the osmotic pressure can be obtained by assuming that the solution contains only sodium chloride. This yields an osmotic pressure of approximately 200 psi for the raw well water, and 272 psi for the average feedwater concentration through the reverse osmosis unit. The average feedwater concentration is calculated as the arithmetic average of the feedwater and brine concentrations. Calculation of the average osmotic pressure is necessary because of the staggered array of the pressure vessels; the brine from vessels 1 and 2 becomes the feedwater for vessel 3.

Since the osmotic pressure tends to draw water away from the low-concentration side of the membrane (product water) toward the high-concentration side (feedwater), it acts against the pressure developed by the system pumps (see Appendix A). To account for this, the net driving pressure is calculated by subtracting the osmotic pressure from the pump pressure. A summary of these values, and their effect on membrane performance, is presented in Table 1. As this table illustrates, 500-psi net driving pressure was not realized, and consequently 50% recovery could not be achieved. The practical limit for the pumps supplied with the demonstration unit was 600 psi, and thus operation at higher pressure was not possible. A reverse osmosis unit designed for permanent installation at Tom's Cove could either be equipped with a pump capable of operating at up to 800 psi (such as a Gould pump) or designed with a pressure vessel array that would allow 50% recovery at 600 psi.

Owing to the limitations imposed by osmotic pressure, a recovery of approximately 50% is probably the maximum desirable recovery on a feedwater such as that found at Tom's Cove. At 50% recovery the calculated average osmotic pressure through the unit would be approximately 330 psi. As the osmotic pressure increases, the energy and/or membrane area required to overcome it becomes excessive, which increases the cost of the product water.

TABLE 1

EFFECT OF OSMOTIC PRESSURE - ASSATEAGUE ISLAND

Date	Pump Pressure (psi)	Osmotic Pressure (psi)	Net Driving Pressure (psi)	Observed GFD*	GFD/Net Pressure (psi)
6/25/70	500	260	240	4.32	0 .018
6/26/70	500	260	240	4.13	0 .017
6/27/70	500	260	240	4.16	0 .017
6/28/70	500	260	240	3.90	0 .016
6/29/70	500	260	240	4.09	0 .017
6/30/70	500	260	240	4.03	0 .017
7/1/70	600	284	316	5.06	0 .016
7/2/70	600	284	316	4.96	0 .016
7/3/70	600	284	316	4.86	0 .015
7/4/70	600	284	316	4.93	0 .016
7/5/70	600	284	316	4.86	0 .016
7/5/70	430	28	402	7.23	0 .013

*Gallons per square foot of membrane per day.

PRETREATMENT AND PRODUCT WATER FLUX

The well water being processed during the 11-day demonstration run had a TDS content of approximately 19,000 mg/l (dried at 105°C). Owing to the nature of this well water, all the pretreatment items provided with the unit were employed, including addition of concentrated sulphuric acid, chlorine, sodium hexametaphosphate, and 5-μ filtration.

Approximately 25 gal of concentrated sulphuric acid was utilized during the demonstration run to maintain the feedwater pH between 5.0 and 6.0. This is equivalent to 1 gal of concentrated sulphuric acid per 2731 gal of raw feedwater. The feedwater pH is adjusted to prevent precipitation of calcium carbonate and to extend the membrane lifetime. At a feedwater pH above 6.0, membrane hydrolysis proceeds at an unacceptably fast rate.

Chlorine was added at a rate of 0.5 mg/l residual to protect the membrane from biological attack. This is necessary because microorganisms commonly occur in natural waters which may damage the membrane surface.

Sodium hexametaphosphate was added at a rate of 5 mg/l to guard against precipitation of sparingly soluble salts such as calcium sulfate. This additive helps to disperse the salt and allows it to be flushed from the unit with the brine stream.

The 5-μ cartridge filters were provided to prevent foreign material from damaging the high-pressure pumps or fouling the modules. The filters were changed every 12 hr during this demonstration run. The major cause of plugging was sand pumped out of the well by the transfer pump (see Fig. 7). A permanent installation could be equipped with a settling tank and sand filter for increased efficiency and reduced maintenance time.

The pretreatment steps described above were effective in that no membrane damage occurred due to salt precipitation or biological attack. This observation is supported by the steady flux exhibited by the unit during the period of operation. A slight flux decline (-0.05 log log slope) resulting from the

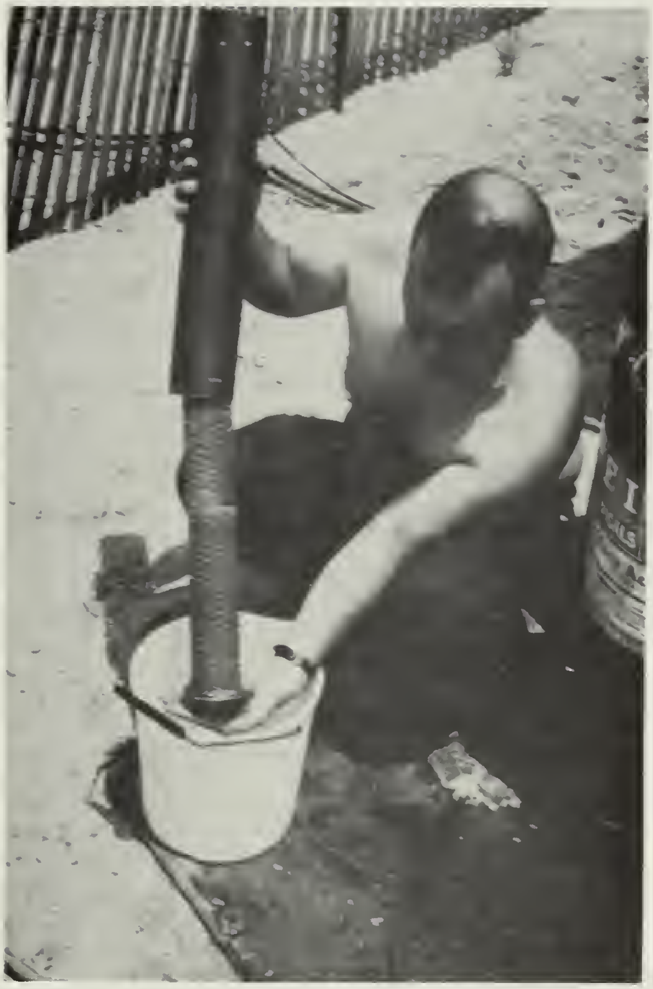

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Fig. 7. Garland Moore, National Park Service, holding sand-clogged cartridge filters

normal membrane property of compaction was anticipated. Over the entire 11-day demonstration run at Tom's Cove, the observed flux decline was even less than that expected to result from compaction alone. Figure 8 is a graph showing the change in the membrane constant (A) with time, along with the anticipated compaction slope. This graph clearly illustrates the lack of a significant flux decline.

MEMBRANE PERFORMANCE

During the 11 days of continuous operation, measurements were taken at least twice daily of pump pressure, flow, pH, water temperature, and conductivity. These data are presented in Tables 2 and 3. In addition, samples of the raw well water; well water after addition of sulphuric acid, chlorine, and sodium hexametaphosphate; product water; and concentrate were collected and returned to San Diego for analysis by an independent laboratory.* The results of these analyses appear in Tables 4 through 9.

As can be seen from Tables 2 and 3, the average salt rejection during this run was 92.9%. The data in Tables 4 through 8 verify that the product water was of significantly improved quality compared with the brackish well water. However, 92.9% rejection was not sufficient to produce potable quality water.

TWO-STAGE OPERATION

It was decided that a second pass through the reverse osmosis unit would be required to produce potable quality water. Subsequently, 100 gal of product water from the first pass through the unit was collected and used as feedwater. The product water produced during the second pass was of excellent potable quality. An analysis of this water is presented in Table 9. A salt rejection of greater than 99% had been achieved in the

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*Environmental Engineering Laboratory, 3467 Kurtz Street, San Diego, California.

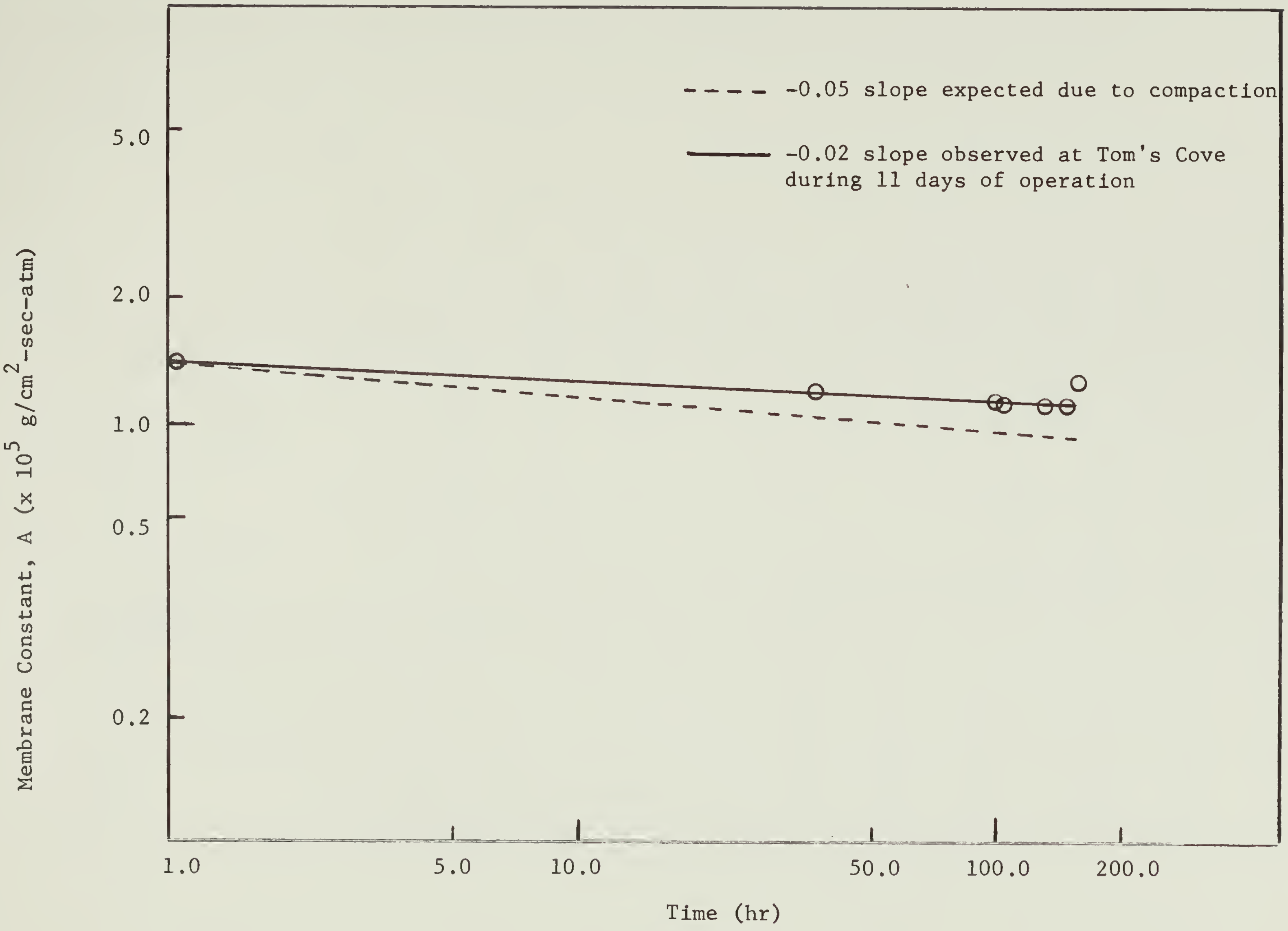

Fig. 8. Membrane constant (A) versus time, and expected compaction slope

TABLE 2

OPERATING DATA - ASSATEAGUE ISLAND (JUNE 25-30, 1970)

Date Time	6/25 14:15	6/25 16:45	6/26 0945	6/26 20:35	6/27 11:37	6/27 20:50	6/28 11:15	6/28 20:15	6/29 09:30	6/29 19:00	6/30 10:15	6/30 19:00
Pressure, psi	500	500	500	500	500	500	500	500	500	500	500	500
Temperature, °F	68	68	68	66	67	66	65	64	66	66	66	65
Feed Conductivity, µmho/cm	25,000	25,000	20,500	27,000	30,000	27,000	26,000	27,000	34,000	36,000	35,000	36,600
Brine Conductivity, µmho/cm	31,500	34,400	26,000	35,000	40,800	38,700	34,500	39,000	46,500	49,500	48,000	49,000
Product Conductivity, µmho/cm	2,300	2,200	1,900	2,200	2,400	2,300	2,900	2,700	3,300	3,300	3,300	3,200
Feed pH	6.7	5.0	5.0	5.0	6.8	6.8	6.8	6.8	5.9	6.2	6.4	5.5
Product Flow, gpm	1.34	1.35	1.29	1.29	1.30	1.28	1.27	1.17	1.28	1.28	1.26	1.29
Brine Flow, gpm	3.98	4.0	4.28	4.28	3.87	3.87	3.87	3.87	3.89	4.0	3.75	3.86
Rejection, %	91.9	92.6	91.8	92.9	93.2	93.0	90.4	91.8	91.8	92.3	92.0	92.5

TABLE 3

OPERATING DATA - ASSATEAGUE ISLAND (JULY 1-5, 1970)

Date Time	7/1 10:00	7/1 18:45	7/2 10:30	7/2 18:00	7/3 09:15	7/3 19:00	7/4 10:00	7/4 19:00	7/5 10:30	7/5 12:30
Pressure, psi	600	600	600	600	600	600	600	600	600	430
Temperature, °F	67	66	67	66	66	66	68	69	65	74
Feed Conductivity, µmho-cm	33,000	35,500	34,000	35,500	33,750	36,000	34,500	34,000	36,500	2,500
Brine Conductivity, µmho/cm	52,000	56,000	50,500	54,000	53,000	55,750	53,750	53,000	53,000	4,080
Product Conductivity, µmho/cm	3,000	2,700	2,800	2,700	2,700	2,600	2,600	2,500	2,600	260
Feed pH	6.7	5.5	5.4	5.0	5.5	5.7	5.15	5.7	5.0	4.3
Product Flow, gpm	1.58	1.59	1.55	1.54	1.51	1.53	1.54	1.54	1.52	2.26
Brine Flow, gpm	3.0	3.0	3.0	3.0	3.09	3.0	3.05	3.01	3.02	2.45
Rejection,%	92.9	94.1	93.4	94.0	93.8	94.3	94.1	94.3	94.2	92.1

TABLE 4

CHEMICAL ANALYSIS - ASSATEAGUE ISLAND

June 19, 1970

(500-psi Operating Pressure)

(Results in mg/l except where noted)

	Raw Well Water	Concentrate	Product	Rejection (%)
Calcium (Ca)	800	880	44	94.8
Magnesium (Mg)	462	656	34	93.9
Sodium (Na)	5,810	6,900	540	91.5
Potassium (K)	164	252	23	88.9
Carbonate (CO_3)	0	0	0	--
Bicarbonate (HCO_3)	647	98	0	100
Sulfate (SO_4)	1,460	1,780	37	97.7
Chloride (Cl)	10,800	12,750	1,180	90.0
Nitrate (NO_3)	0.09	0.22	--	--
Boron (B)	1.0	1.0	0.86	14.0
Silica (SiO_2)	48	60	5.2	90.4
Iron (Fe)	0.07	0.19	--	--
Manganese (Mn)	0	0	0	--
Total Dissolved Solids:				
105°C	19,300	25,000	1,900	91.4
185°C	18,800	23,300	1,850	91.2
Total Alkalinity ($CaCO_3$)	530	80	0	100
Total Hardness ($CaCO_3$)	3,900	4,900	350	92.0
pH	8.0	7.1	2.6*	
Specific Conductivity				
µmho/cm at 25°C	35,000	40,000	3,400	90.9

---

*The pH controller was not operating correctly.

TABLE 5

CHEMICAL ANALYSIS - ASSATEAGUE ISLAND

June 23, 1970

(500-psi Operating Pressure)

(Results in mg/l except where noted)

	Raw Well Water	Concentrate	Product	Rejection (%)
Calcium (Ca)	840	1800	32	97.6
Magnesium (Mg)	437	170	34	88.8
Sodium (Na)	5400	7200	820	87.0
Potassium (K)	200	228	30	86.0
Carbonate (CO_3)	0	0	0	--
Bicarbonate (HCO_3)	647	244	55	87.8
Sulfate (SO_4)	545	1800	42	96.4
Chloride (Cl)	10,700	13,800	1390	88.7
Nitrate (NO_3)	0.04	0.27	--	--
Boron (B)	1.0	1.1	0.79	24.8
Silica (SiO_2)	54	64	4.4	92.5
Iron (Fe)	0.13	0.06	--	--
Manganese (Mn)	0	0	0	--
Total Dissolved Solids:				
105°C	20,400	25,800	2380	89.7
185°C	18,700	24,600	2350	94.6
Total Alkalinity ($CaCO_3$)	530	200	45	87.7
Total Hardness ($CaCO_3$)	3,900	5,200	320	93.0
pH	8.1	8.1	7.1	
Specific Conductivity, µmho/cm at 25°C	31,000	42,000	4050	88.9

TABLE 6

CHEMICAL ANALYSIS - ASSATEAGUE ISLAND

June 27, 1970

(500-psi Operating Pressure)

(Results in mg/l except where noted)

	Raw Well Water	Concentrate	Product	Rejection (%)
Calcium (Ca)	440	608	4.8	99.1
Magnesium (Mg)	632	846	7.3	99.0
Sodium (Na)	6,200	8,000	580	91.8
Potassium (K)	200	260	24	89.6
Carbonate (CO_3)	0	0	0	--
Bicarbonate (HCO_3)	573	756	21	96.8
Sulfate (SO_4)	763	1020	0	100
Chloride (Cl)	11,325	15,150	855	93.5
Nitrate (NO_3)	1.2	1.6	0	100
Boron (B)	1.3	1.5	1.1	21.4
Silica (SiO_2)	52	70	2.0	96.7
Iron (Fe)	8.4	20	0.03	99.8
Manganese (Mn)	0	0	0	--
Total Dissolved Solids:				
105°C	19,330	25,340	1,464	93.4
185°C	17,840	23,640	1,404	93.2
Total Alkalinity ($CaCO_3$)	466	620	17	96.7
Total Hardness ($CaCO_3$)	3,700	5,000	42	99.0
pH	7.1	7.6	6.6	
Specific Conductivity, μmho/cm at 25°C	35,000	46,000	2,500	93.8

TABLE 7

CHEMICAL ANALYSIS - ASSATEAGUE ISLAND

June 30, 1970

(500-psi Operating Pressure)

(Results in mg/l except where noted)

	Raw Well Water	Well Water Plus Additives	Concentrate	Product	Rejection (%)
Calcium (Ca)	420	440	560	14	97.2
Magnesium (mg)	631	632	851	19	97.4
Sodium (Na)	6,400	6,200	8,000	660	90.7
Potassium (K)	200	200	260	28	87.8
Carbonate (CO_3)	0	0	0	0	--
Bicarbonate (HCO_3)	647	39	110	24	67.8
Sulfate (SO_4)	832	1,380	1,720	0	100
Chloride (Cl)	11,400	13,000	15,000	1,060	92.4
Nitrate (NO_3)	0.08	0.08	0.10	0	100
Boron (B)	1.4	1.3	1.5	1.1	21.4
Silica (SiO_2)	36	34	49	3.4	91.8
Iron (Fe)	19	18	24	0.11	99.5
Manganese (Mn)	0	0	0	0	--
Total Dissolved Solids:					
105°C	19,350	19,970	25,550	1,820	92.0
185°C	18,710	18,080	23,970	1,728	91.8
Total Alkalinity ($CaCO_3$)	530	32	90	20	67.5
Total Hardness ($CaCO_3$)	3,650	3,700	4,900	112	97.4
pH	7.6	5.8	6.4	5.4	
Specific Conductivity, µmho/cm at 25°C	35,000	35,000	45,000	3,100	92.3

TABLE 8

CHEMICAL ANALYSIS - ASSATEAGUE ISLAND

July 5, 1970

(600-psi Operating Pressure)

(Results in mg/l except where noted)

	Raw Well Water	Well Water Plus Additives	Concentrate	Product	Rejection (%)
Calcium (Ca)	420	408	640	4.8	99.1
Magnesium (Mg)	632	639	948	6.3	99.2
Sodium (Na)	6,200	6,400	9,200	600	92.3
Potassium (K)	200	200	280	24	90.0
Carbonate (CO_3)	0	0	0	0	--
Bicarbonate (HCO_3)	634	122	183	12	92.1
Sulfate (SO_4)	842	1,208	1,889	0	100
Chloride (Cl)	11,300	11,100	16,700	860	93.8
Nitrate (NO_3)	0.08	0.08	0.10	0	100
Boron (B)	1.4	1.4	1.6	1.1	26.7
Silica (SiO_2)	34	34	46	2.4	94.0
Iron (Fe)	11	14	21	0.14	99.2
Manganese (Mn)	0	0	0	0	--
Total Dissolved Solids:					
105°C	19,380	19,990	29,330	1,420	94.2
185°C	18,070	18,330	27,060	1,368	94.0
Total Alkalinity ($CaCO_3$)	520	100	150	10	92.0
Total Hardness ($CaCO_3$)	3,650	3,650	5,500	38	99.2
pH	7.4	6.1	6.3	4.9	
Specific Conductivity, μmho/cm at 25°C	34,500	34,500	51,500	2,600	94.0

TABLE 9

CHEMICAL ANALYSIS - ASSATEAGUE ISLAND

July 5, 1970

(430-psi Operating Pressure)

(Results in mg/l except where noted)

	Product Water Used as Feed Water	Concentrate*	Product	Rejection (%)
Calcium (Ca)	4.8	40	0	100
Magnesium (Mg)	6.3	22	0	100
Sodium (Na)	600	1120	56	93.5
Potassium (K)	24	40	2.6	91.9
Carbonate (CO_3)	0	0	0	--
Bicarbonate (HCO_3)	12	0	11	--
Sulfate (SO_4)	0	417	0	--
Chloride (Cl)	860	1800	70	94.7
Nitrate (NO_3)	0	0	0	--
Boron (B)	1.1	-	0.84	23.6
Silica (SiO_2)	2.4	16	0.4	95.7
Iron (Fe)	0.14	1.3	0.07	90.2
Manganese (Mn)	0	0	0	--
Total Dissolved Solids:				
105°CC	1,420	3476	156	93.6
185°CC	1,368	-	104	92.4
Total Alkalinity ($CaCO_3$)	10	0	9	--
Total Hardness ($CaCO_3$)	38	190	0	100
pH	4.9	2.7	4.9	
Specific Conductivity				
μmho/cm at 25°CC	2,600	6000	240	94.4

*This sample was not taken at the same time as the other two, and in the interim a portion of sulphuric acid was accidentally added, causing the unusually low pH and incorrect sulfate and bicarbonate readings.

conversion from the well water concentration of 19,000 mg/l to the product water concentration after the second pass of 156 mg/l.

The only treatment the product water from the second pass would require before it could be utilized as a potable water supply would be pH adjustment. The pH of the product water was 4.9 and should be adjusted to approximately 7.0. Carbon dioxide removal by decarbonization or degasification should increase the product water pH significantly. In addition, the product water could be treated with either limestone or caustic in small enough quantities for effective further neutralization, without interfering with the taste. The exact amount of chemicals required to neutralize the product is dependent on the effectiveness of decarbonization and can best be determined empirically. The economics of product water neutralization are discussed in Appendix C.

The effective recovery of the second stage of a two-stage unit could be established at 100% by blending all the second-stage brine with the raw feedwater entering the first stage. The brine observed after the second pass at Tom's Cove had a TDS concentration of approximately 3500 mg/l. Blending it with the 19,000 mg/l raw feedwater would actually be beneficial in that the raw feedwater would be diluted slightly. Operation in this manner would establish the total recovery of a two-stage unit as that recovery achieved by the first stage. With the recovery of the first stage set at 50%, the total system recovery would be 50%.

4. SUMMARY OF UNIT PERFORMANCE AT TOM'S COVE

A continuous 11-day reverse osmosis demonstration run was completed on a well water feed containing approximately 19,000 mg/l total dissolved solids. The reverse osmosis unit was operated at 500-psi pump pressure for the first 6 days and 600 psi for the remainder of the test. Owing to osmotic pressure, the net driving pressures were actually 240 and 312 psi, respectively, during these periods. Sulphuric acid, chlorine, and sodium hexametaphosphate were added to the well water before exposure to the membrane. In addition, 5-μ cartridge filtration was employed.

During the 11-day run the membrane rejected an average of 92.9% of the dissolved solids in the brackish well water. From a TDS concentration of 19,000 mg/l, the well water was reduced to 1400 mg/l TDS in the product. The product water flux was essentially constant for the entire 11 days, indicating that fouling was not taking place.

At the conclusion of the 11-day run, 100 gal of product water was collected and passed through the unit a second time, simulating a two-stage system. The product water after the second pass had a TDS content of 156 mg/l. The only treatment required before the product water could be used as a potable water source would be pH adjustment.

5. CONCLUSIONS

Based on the information gathered during this field demonstration, the following conclusions can be drawn:

1. A two-stage reverse osmosis unit operating on a feedwater such as that found at Tom's Cove would be capable of producing excellent quality potable water. This unit could operate at 50% recovery.

2. No significant product water flux decline or membrane attack should occur if the pretreatment described in this report is employed.

3. Product water post-treatment would be necessary but would only consist of decarbonization followed by minimal chemical addition.

APPENDIX A

REVERSE OSMOSIS: A GENERAL DESCRIPTION

Osmosis can be defined as the spontaneous passage of a liquid from a dilute to a more concentrated solution across a semipermeable membrane. This membrane allows the passage of the solvent (water) but not the dissolved solids (solutes) (see Fig. 9). The transfer of the water from one side of the membrane to the other continues until the pressure is large enough to prevent any net transfer of the water to the more concentrated solution. At equilibrium, the quantity of water passing in either direction is equal, and the pressure is then defined as the osmotic pressure of the solution having that particular concentration of dissolved solids.

Reverse osmosis is a process achieved by placing a pressure on the concentrated solution (by means of a pump) sufficient to overcome the osmotic pressure. Thus, the water is driven from the more concentrated solution, through the semipermeable membrane, and into the less concentrated side (see Fig. 10).

The cellulose acetate semipermeable membrane used in almost all reverse osmosis systems acts as a highly effective barrier to the passage of dissolved and suspended substances. Gulf Environmental Systems reverse osmosis units utilize a modified cellulose acetate membrane capable of rejecting up to 96.5% sodium chloride at 600 psi.

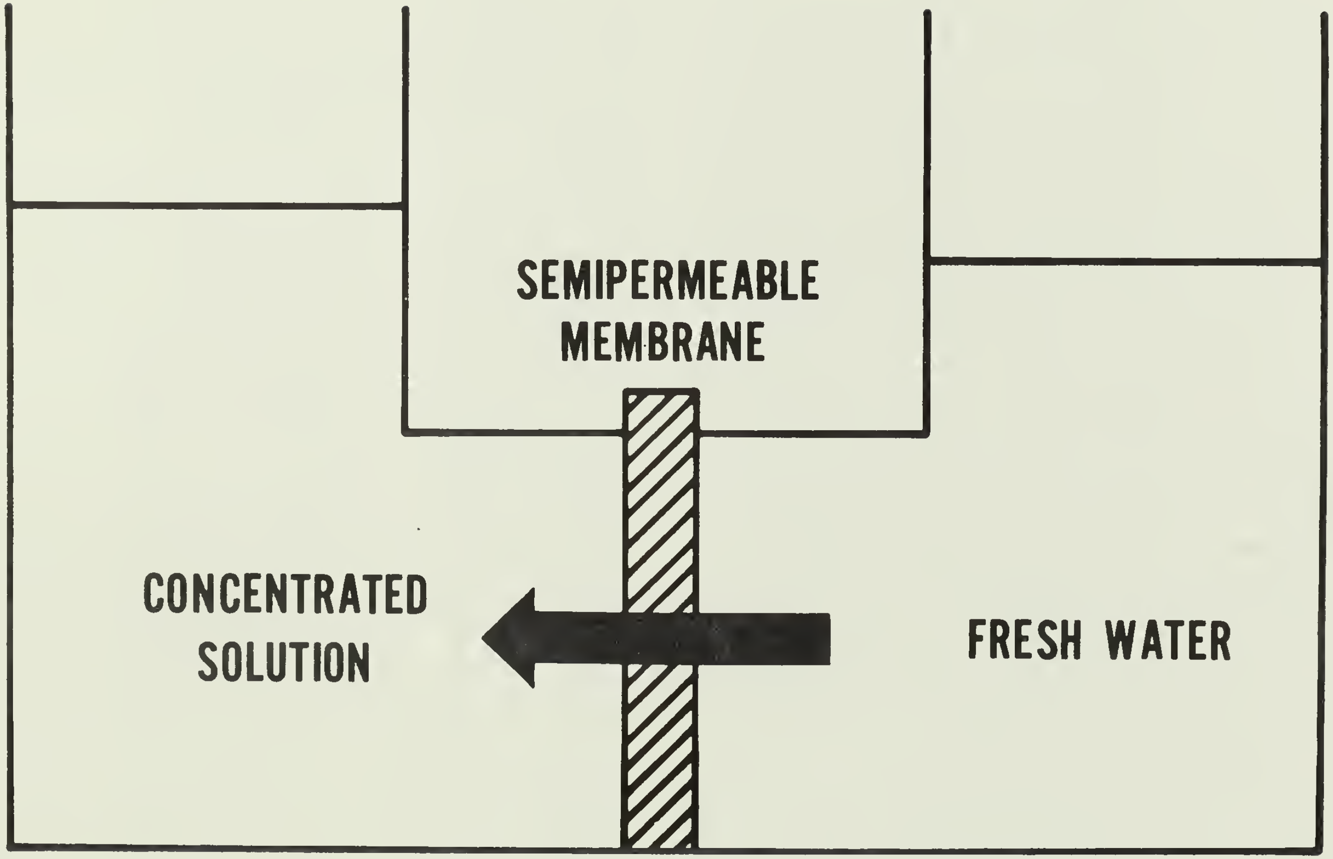

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Fig. 9. Osmosis: normal flow from low to high concentration

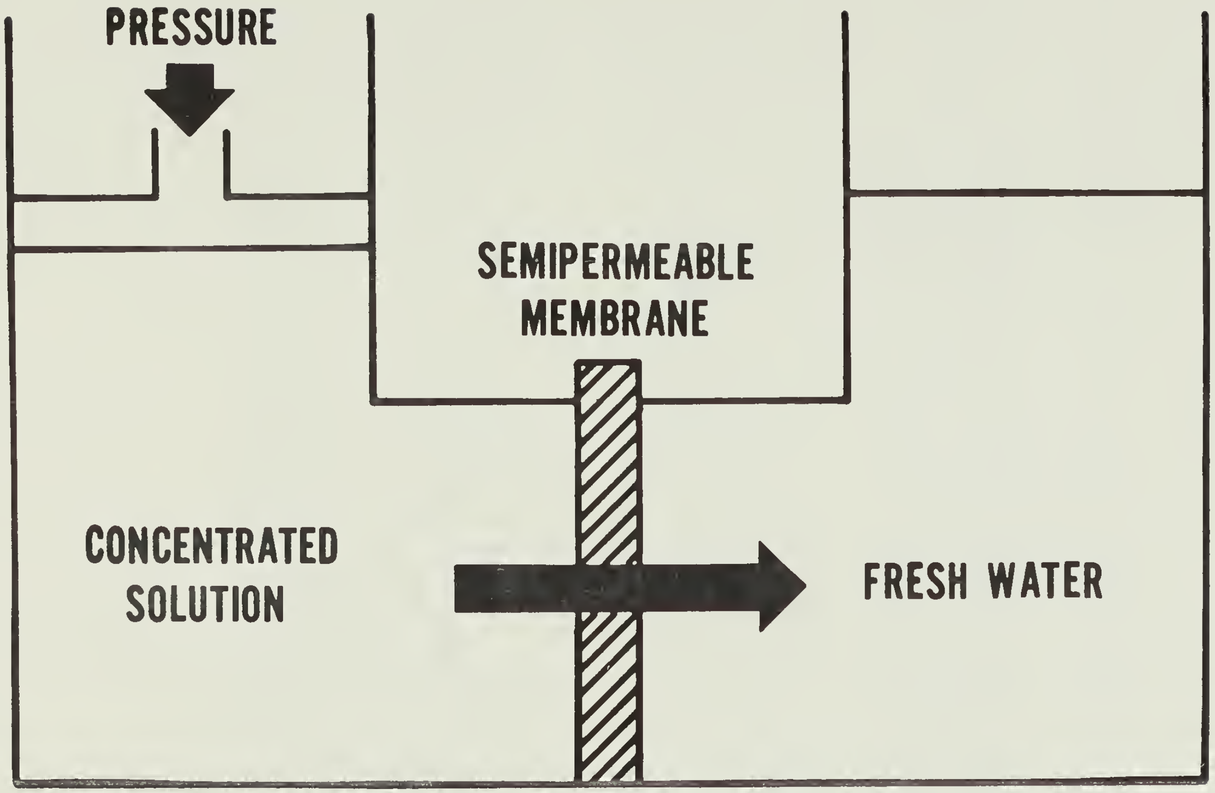

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Fig. 10. Reverse osmosis: flow reversed by application of pressure to high-concentration solution

APPENDIX B

U.S. PUBLIC HEALTH SERVICE DRINKING WATER STANDARDS*

Chemical substances listed in Table 10 should not be present in a water supply in excess of the listed concentrations where, in the judgment of the Reporting Agency and the Certifying Authority, other more suitable supplies are, or can be, made available.

When fluoride is naturally present in drinking water, the concentration should not average more than the appropriate upper limit in Table 10. Presence of fluoride in average concentrations greater than two times the optimum values in Table 11 shall constitute grounds for rejection of the supply.

Where fluoridation (supplementation of fluoride in drinking water) is practiced, the average fluoride concentration shall be kept within the upper and lower control limits in Table 11.

Presence of the substances listed in Table 12 in excess of the concentrations listed shall constitute grounds for rejection of the supply.

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*U.S. Department of Health, Education, and Welfare, Public Health Drinking Water Standards, 1962, U.S. Government Printing Office, Washington, D.C., pp. 7 and 8.

TABLE 10

MAXIMUM CHEMICAL SUBSTANCE LIMITS

Substance	Concentration (mg/l)
Alkyl Benzene Sulfonate (ABS)	0. 5
Arsenic (As)	0. 01
Chloride (Cl)	250. 0
Copper (Cu)	1. 0
Carbon Chloroform Extract (CCE)	0. 2
Cyanide (CN)	0. 01
Fluoride (F)	(See Table 11)
Iron (Fe)	0. 3
Manganese (Mn)	0. 05
Nitrate (NO_3)	45. 0
Phenols	0. 001
Sulfate (SO_4)	250. 0
Total Dissolved Solids	500. 0
Zinc (Zn)	5. 0

TABLE 11

FLUORIDE LIMITS

Annual Average of Maximum Daily Air Temperatures (°F)	Recommended Control Limits, Fluoride Concentrations (mg/l)		
	Lower	Optimum	Upper
50.0 to 53.7	0.9	1.2	1.7
53.8 to 58.3	0.8	1.1	1.5
58.4 to 63.8	0.8	1.0	1.3
63.9 to 70.6	0.7	0.9	1.2
70.7 to 79.2	0.7	0.8	1.0
79.3 to 90.5	0.6	0.7	0.8

TABLE 12

MAXIMUM SUBSTANCE LIMITS

Substance	Limit (mg/l)
Arsenic (As)	0.05
Barium (Ba)	1.0
Cadmium (Cd)	0.01
Chromium (Hexavalent) (Cr^{+6})	0.05
Cyanide (CN)	0.2
Fluoride (F)	(See Table 11)
Lead (Pb)	0.05
Selenium (Se)	0.01
Silver (Ag)	0.05

APPENDIX C

ECONOMICS OF REVERSE OSMOSIS

When the decision was reached to conduct a reverse osmosis field demonstration in the Tom's Cove area of Assateague Island, it was believed that the brackish well water there had a TDS content of less than 5000 mg/l. By operating on a raw water supply of this concentration (assuming it contains no large quantities of poorly rejected ions, such as boron), potable water could be produced in one pass. However, since the well water characterized in Tables 4 through 9 of this report may be the best available raw water source at Assateague Island, a two-stage reverse osmosis unit is apparently required for potable water production. Two-stage units necessitate additional pumping capability and a larger number of modules per gallon of product water produced, as compared with one-pass units. The economics for two-stage reverse osmosis units are presented in this appendix.

ECONOMICS OF A TWO-PASS REVERSE OSMOSIS UNIT

Gulf Environmental Systems has designed a 5000-gpd two-pass reverse osmosis unit and tested it on Pacific Ocean water. The experience gained operating this unit, combined with the results of laboratory scale tests, makes it possible to present realistic cost estimates for two-stage units.

The following economics apply to two-stage reverse osmosis units employing either standard Model 4000 modules (such as those used at Tom's Cove) or high-flux modules. The high-flux modules initially have twice the flux capability of the Model 4000 modules. However, high-flux membrane is a relatively new development, and the high-flux modules might require more frequent replacement.

The 5000-gpd seawater test unit employed both Model 4000 modules and high-flux modules. The economics presented using either module are thus based on actual operating experience.

OPERATION OF A TWO-STAGE UNIT

The first stage of a two-stage reverse osmosis unit would be operated at between 600 and 800 psi and approximately 50% recovery. The product water concentration from this pass would be from 1000 to 1500 mg/l TDS.

The second stage would operate at from 300 to 600 psi and 100% recovery. This can be accomplished by recycling the brine and blending it with the raw feedwater entering the first stage. This technique would yield a total system recovery rate equal to the recovery of the first stage, or approximately 50%.

OPERATING COSTS

The operating costs of a two-stage reverse osmosis unit can be divided into four basic categories: power, chemicals, operation and maintenance, and module replacement.

Power

The primary energy requirements for a reverse osmosis unit are associated with the high-pressure pumps. Experience indicates that the ancillary equipment requirements (transfer pumps, booster pumps, chemical feeders, etc.) will add approximately 10% to this load. Assuming 1 kWh/1000 gal feed/100 psi, and 1¢ per kWh, the following power requirements can be calculated:

1st stage at 50% recovery - 800 psi = 16 kWh/1000 gal product
2nd stage at 80% recovery* - 600 psi = 7.5 kWh/1000 gal product
10% allowance for auxiliaries = 2.4 kWh/1000 gal product
Total = 26 kWh/1000 gal product

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*All the feedwater going to the second stage will be utilized in that the 20% brine will be recycled and blended with the raw feedwater.

Chemicals

Chemical treatment of the feedwater will require adjustment of the pH by acid addition to eliminate possible carbonate scaling and to improve module lifetime. A threshold treatment such as sodium hexametaphosphate will also be added to control sulfate scaling. Chlorination will be necessary to control organic sliming or fouling. Furthermore, it will be necessary to decarbonate and treat the product with either limestone or caustic to restore the pH to near neutral.

It is usually possible to avoid carbonate scaling by neutralizing three-fourths of the bicarbonate alkalinity. On this basis, the cost of neutralizing 400 mg/l bicarbonate alkalinity (as $CaCO_3$) using sulphuric acid will range downward from 9¢ to 7¢/1000 gal of feed. Sodium hexametaphosphate addition at a rate of 5 mg/l will cost approximately 0.75¢/1000 gal of feed. Chlorination, to achieve a residual of 0.5 to 1.0 mg/l, will be a trivial cost increment. The chemicals necessary for product water neutralization, when used to augment decarbonization, will cost about one-tenth as much as acidification of the raw feedwater. Chemical costs can be summed up as follows:

pH adjustment, (7¢/1000)/0.50 = 14¢/1000 gal product
Threshold treatment, (0.75/1000)/0.50 = 1.5¢/1000 gal product
Product neutralization, 1.4¢/1000 gal product
Total = 17.0¢/1000 gal product

Operation and Maintenance

Operating experience with reverse osmosis systems to date indicates that operation and maintenance labor requirements are minimal, and that maintenance material requirements will be quite low. The estimates presented below are based on 330 operating days per year. Five percent of the cost of non-module components as an annual maintenance charge must be added to these figures.

	Unit Size (1000 gpd)			
	4	10	25	100
Manpower at $15,000/yr	0.25	0.33	0.66	2

Module Replacement

The membrane lifetime for a two-stage reverse osmosis unit should not be markedly different than that for a one-pass unit. Experience indicates that a 3-yr module lifetime is a reasonable assumption.

Capital Costs

The capital costs presented in this appendix include the reverse osmosis unit (f.o.b. San Diego), a pretreatment skid which provides safety filtration for the pump(s) and modules, an automatic pH controller, an acid feed pump, a chemical feed pump, a booster pump, and a control system adequate for the application. The figures presented are based on 70° to 77°F water temperature. An amortization rate of 8% annually is included based on the capital costs.

Water Costs

Water costs are presented as a function of plant size and reflect the total costs anticipated per 1000 gal of product water. These costs include all the factors discussed above and should facilitate a cost comparison between reverse osmosis and other means of acquiring potable water.

The water costs for two-stage reverse osmosis units using Model 4000 modules are presented in Table 13. The water costs for two-stage reverse osmosis units using high-flux modules are presented in Table 14.

TABLE 13

ESTIMATE OF COSTS FOR TWO-STAGE REVERSE OSMOSIS UNITS USING MODEL 4000 MODULES

Size of Unit (gpd)	4,000	10,000	25,000	100,000
Power	0.26	0.26	0.26	0.26
Chemicals	0.17	0.17	0.17	0.17
Operation	2.00	1.37	1.07	0.60
Maintenance	0.52	0.35	0.18	0.10
Membrane	0.91	0.91	0.91	0.70
Subtotal	3.86	3.06	2.59	1.83
Capital @ 8% Amortization Annually	1.76	1.20	0.75	0.57
Total $/1000 gal	5.62	4.26	3.34	2.40
Capital Cost, $1,000	36.3	49.3	79	234

TABLE 14

ESTIMATE OF COSTS FOR TWO-STAGE REVERSE OSMOSIS UNITS USING HIGH-FLUX MODULES

Size of Unit (gpd)	4,000	10,000	25,000	100,000
Power	0.26	0.26	0.26	0.26
Chemicals	0.17	0.17	0.17	0.17
Operation	2.00	1.37	1.07	0.60
Maintenance	0.40	0.27	0.14	0.08
Membrane	0.46	0.46	0.46	0.35
Subtotal	3.29	2.53	2.10	1.46
Capital @ 8% Amortization Annually	1.50	0.98	0.59	0.38
Total $/1000 gal	4.79	3.51	2.69	1.84
Capital Costs, $1,000	31	40	62	157